Advanced O_2 Separation System Integration for Conceptual Design of Supercritical O_2-Based PC Boiler

Topical Report – Rev. 1

Zhen Fan
Andrew Seltzer

June 2006

DE-FC26-04NT42207

Task 2

Foster Wheeler Power Group, Inc.
12 Peach Tree Hill Road
Livingston, New Jersey 07039

DISCLAIMER

ABSTRACT

The objective of the advanced oxygen separation system integration task of the Conceptual Design of Oxygen-Based Supercritical PC Boiler study is to evaluate the benefits, effects, and limitations of the integration of advanced oxygen separation technologies into a supercritical O_2-fired PC. Simulations of the power generation unit, oxygen separation unit, and CO_2 sequestration system were conducted using the Aspen Plus software. The improvement of the O_2-fired PC system performance incorporating the Oxygen Ion Transport Membrane (OITM) and Ceramic Auto-thermal Recovery (CAR) were investigated. A parametric study was conducted to determine the sensitivity of the design and performance to various variables. Compared to the other CO_2 removal and sequestration technologies, the oxygen-fired PC integrated with OITM shows substantially less CO_2 removal penalty. The CO_2 removal penalty of the oxygen-fired PC integrated with CAR appears to be midway between cryogenic air separation and OITM.

Table of Contents

List of Figures

List of Tables

1.0 Introduction

The objective of this task is to identify promising, low cost advanced options of oxygen separation, which can be integrated into the O_2-fired PC power plant. Currently, a number of new oxygen separation technologies are in development. They are categorized as high temperature ion membranes, such as oxygen ion transport membrane (OITM), or high temperature sorption, such as ceramic auto-thermal recovery (CAR). The former relies on oxygen transport through a ceramic membrane, and the latter on oxygen storage in perovskite type materials. Integration of these advanced oxygen separations into the O_2-fired PC has the potential to substantially reduce the cost of CO_2 removal. This task deals with the system-level evaluation of the O_2-fired PC integrated with these advanced oxygen separation methods.

The advanced oxygen separation system integration task, which was performed using the Aspen Plus computer program, is aimed at a system level optimization to minimize the overall heat rate and maximize system performance. Two types of advanced oxygen separation systems and related configurations were simulated: 1) high temperature membrane technology (OITM) and 2) high temperature oxygen sorbent technology (CAR). Determined are the required performance characteristics of the operating components such as the boiler (with air heater for OITM), GT expander, wet-end economizer for low-grade heat recovery, and air separation equipment.

2.0 Executive Summary

The objective of this task (Task 2) in the Conceptual Design of Oxygen-Based Supercritical PC Boiler Study is to develop a system design of a conceptual pulverized coal-fired oxygen combustion power plant, integrated with advanced oxygen separation technology. Using the Aspen Plus computer program, the PC boiler plant is optimized by minimizing the CO_2 removal penalty.

The air-fired reference plant employs a supercritical steam turbine with conditions, 4035 psia/1076ºF/1112ºF/2.0”Hg, fires high-volatile bituminous coal, and produces 460 MW at the generator. A conventional air-fired case was simulated as the comparison basis. The air-fired plant has a net plant efficiency of 39.5% and a net power generation of 430 MWe.

The O_2-fired reference plant with cryogenic ASU has a net plant efficiency of 31.9%, a net power generation of 338 MWe, and a CO_2 removal penalty of 114 kWh/klbCO_2.

Oxygen separation by oxygen ion transport membrane is driven by the difference in oxygen partial pressure across a membrane. To produce this pressure difference, the air is pressurized by a compressor and heated to a high temperature. The hot pressurized air is fed to OITM, where about 85% of its O_2 is separated through membrane, and the rest of hot vitiated air is sent to a gas expander. Power generated from the expander is used to drive the air compressor. The OITM does not consume electrical power; instead, it absorbs heat, generates power, as it separates O_2 from air.

The O_2-fired reference plant with OITM has a net plant efficiency of 36.1%, a net power generation of 463 MWe, and a CO_2 removal penalty of 42 kWh/klbCO_2.

Parametric trade-off runs were conducted by varying the O_2 recovery efficiency, the pressure difference across the membrane, the OITM operating pressure, the compressor discharge temperature, and the furnace flame temperature.

The ceramic auto-thermal recovery (CAR) process is based on sorption and storage of oxygen in a fixed bed containing ionic and electronic conductor materials. For the CAR process when extracted steam is used as sweep gas, it boosts net system efficiency only by 0.7% point compared to the cryogenic ASU process. But if the CAR process uses recycle flue sweep gas, system efficiency can be increased by 2.6% points, compared to the gain of 3.2% points of the OITM process.

The O_2-fired PC CO_2 removal penalty with integration of OITM is nearly a quarter of that from post combustion CO_2 removal technologies, and only a half of IGCC. OITM faces significant challenges with respect to the manufacture and stability of membranes, and scale up and design of large plants.

3.0 Experimental

This work performed for this report was performed utilizing computer program simulations. No experimental equipment was used.

4.0 Results and Discussion

4.1 Reference Site and Conditions

In December 2000, Parsons published a study of the cost of electricity of several case studies in CO_2 sequestration from a PC boiler by post capture [1]. In September 2005, Foster Wheeler released a report of conceptual design of O_2-fired PC boiler [3]. To provide a consistent comparison with the cases analyzed in the previous reports, the same site conditions (59ºF, 14.7 psia, 60% RH) and the same fuel (Illinois #6) were used. Site Conditions and fuel properties are presented in Table 1. Fuel HHV and LHV were estimated by a DuLong's method and the stoichiometric air ratio of 867 lb_{air}/lb_{coal} was calculated based on the fuel ultimate analysis.

Table 1 - Site Condition

Standard site:			**air, %v**	dry	wet		**coal, Ill#6**	%w		**sorb**	%w
elevation, ft	0		N2	78.085	77.297		C	63.75		CaCO3	100
amb p, psia	14.70		O2	20.947	20.735		H	4.5			
amb T, F	59		Ar	0.935	0.926		O	6.88			
amb T, wet, F	51.5		CO2	0.033	0.033		N	1.25			
RH, %	60		H2O	0.000	1.010		S	2.51			
P-H2O, psia	0.247		sum	100.000	100.000		A	9.99			
Y-H2O, %v	1.010						M	11.12			
condenser P, "Hg	2.00						V	34.99			
							F	44.19			
							sum	100.0			
							fuel HHV	btu/lb			
							given	11666			
							aspen	11631			

The CO_2 fluid produced from the oxygen-based PC power plant is not chemically pure, but can readily sequestered in geologic formations (depleted oil and gas reservoirs, unmineable coal seams, saline formations, and shale formations) or in oceans. The liquid CO_2 exits the plant at over 2000 psia. The CO_2 fluid inside the pipeline under this pressure is in a liquid or a supercritical state. The other gases in the delivered CO_2 are limited to H_2O < 50 ppm (to avoid acid corrosion), and $Ar+N_2$ < 3% (to avoid phase separation). The excess gases in CO_2 stream either have to be purged or recycled. However, since SO_2, as an acid gas, similar to CO_2, it can be sent to pipeline directly under moisture free condition, and as mentioned in literature, it does not need to be separated out from CO_2 product. Furthermore it is also not necessary to remove the small concentration of NOx in the CO_2 effluent since it can be sequestered along with the CO_2.

4.2 Air-Fired Reference Case (case-1)

To study the effects of CO_2 removal on the performance of power plant, an air-fired supercritical PC boiler was been simulated in detail as a reference case. The reference plant employs a supercritical steam turbine with conditions, 4035 psia/1076ºF/1112ºF/2.0"Hg, fires high-volatile bituminous coal, and produces 460 MW at the generator. It employs eight feed water heaters to raise the final feed water temperature to 569ºF. It uses an auxiliary steam turbine to directly drive the high-pressure feed water pump. Case-1 is the reference air-fired supercritical PC boiler case, and the model and results shown in Figure 1 and Table 2.

Table 2 - Setup and Assumptions

Setup and Result							
ST		**Result**		**Aux power**	MWe	**dP-air**	"H2O
main P, psia	4035	net power, MWe	430	condensed water pump	0.6	AAHX	4.0
main T, F	1076	net eff, %	39.5	HP feed water pump	17.5	duct	6.0
RH P, psia	823	gross @ST, MW	477	FGD pump	0.8	nozzle	10.0
RH T, F	1112	aux power, MW	46.0	CT pump	3.4	sum	20.0
FWHs	8	as %	9.7	PA Fan	1.1		
end wet, %	9.5	HHV in, mmbtu	3721	SA Fan	2.0	**dp-gas**	"H2O
end P, "Hg	2.0	Q to st, mmbtu	3281	ID Fan	5.4	FSH	0.5
		Q, cond, mmbtu	1696	FGD Fan	4.1	FRH	0.5
FWH	F	boiler eff, %	88.2	cooling tower Fan	1.9	RH	1.0
TD	5	ST cycle eff, %	47.7	coal handling	2.0	PSH	0.7
DC	10	Generator eff, %	98.3	sorb handling	0.8	UECO	0.3
FW T	569			ash handling + ESP	1.8	ECO	2.0
		Flow		others (=1%)	4.6	AAHX	1.6
DeSuperheat		air, klb	3270	total	46.0	Damper	4.3
SH, %	5	coal, klb	319			BHG	5.5
water T, F	569	sorb, klb	26	**FGD & SCR**		FGD	12.0
		flue gas, klb	3556	L/G	10	sum	28.4
Boiler		O2, %	3.0	Ca/S	1.05		
PA, %	20	H2O, %	8.6	Excess air, %	85	**dP-Fan**	"H2O
UBC, %	1.0	CO, ppmv	14	NH3/NOx	1.0	PAFan	60
radiation/margin, %	0.59	NOx, ppmv	22	DeSOx, %	98	IDFan	28
EXA, %	17.9	SOx, ppmv	1979	DeNOx, %	90	SAFan	20
flame T, F	3685	after FGD	37				
stack T, F	289	Ash, klb	33			**eff**	%
blowdown, %	0	C, %	6.0			FDFan	75
miller exit T, F	219	main st, klb	2950			IDFan	70
		RH st, klb	2406			CWPump	80
		end st, klb	1573			BFPump	80
						Motor/mech	95

The Aspen Plus model includes coal mills, flue gas heater, pulverized coal-fired furnace, steam generator, superheater, reheater, economizer, ash-removal unit, nitrogen oxides (NOx) selective catalytic reactor (SCR), flue gas de-sulfurization reactor (FGD), air blower, induced draft (ID) fan, feed water pump, cooling water pump, feed water heaters, and a single reheat steam turbine. A coal drying function has been modeled and added into mill module to produce the correct mill exit gas temperature. The furnace was simulated by a zero dimensional model for heat and mass balances. However, all key tube banks of the heat recovery area (HRA) were individually modeled. The furnace roof heat absorption was also simulated. The high-pressure steam temperature is controlled by water spray for de-superheat. The simulation also included heat losses from boiler and HRA enclosure, as well as from the steam pipes. Some user-defined models were included to perform emission calculations. User built-in calculations were added to determine boiler efficiency, system net efficiency, and net power. The heat carried by the exhaust streams was calculated by the program.

The system configuration, detailed setup parameters and summary of results for the case-1 reference case are shown in Figure 1 and Table 2. The system has a steam turbine cycle efficiency (generator power divided by heat transferred to the steam cycle) of 47.6%, a boiler efficiency (heat to steam cycle divided by heat input from fuel to boiler) of 88.2%, an unburned carbon loss (UBC) of 1.0%, and a net plant efficiency of 39.46% (net plant heat rate of 8647 Btu/kWh). It has a gross power of 460 MW at the generator, 18 MW from auxiliary steam expander, an auxiliary power of 46 MW, and a net power of 430 MW. Total heat input from the fuel is 3720 MM Btu/hr.

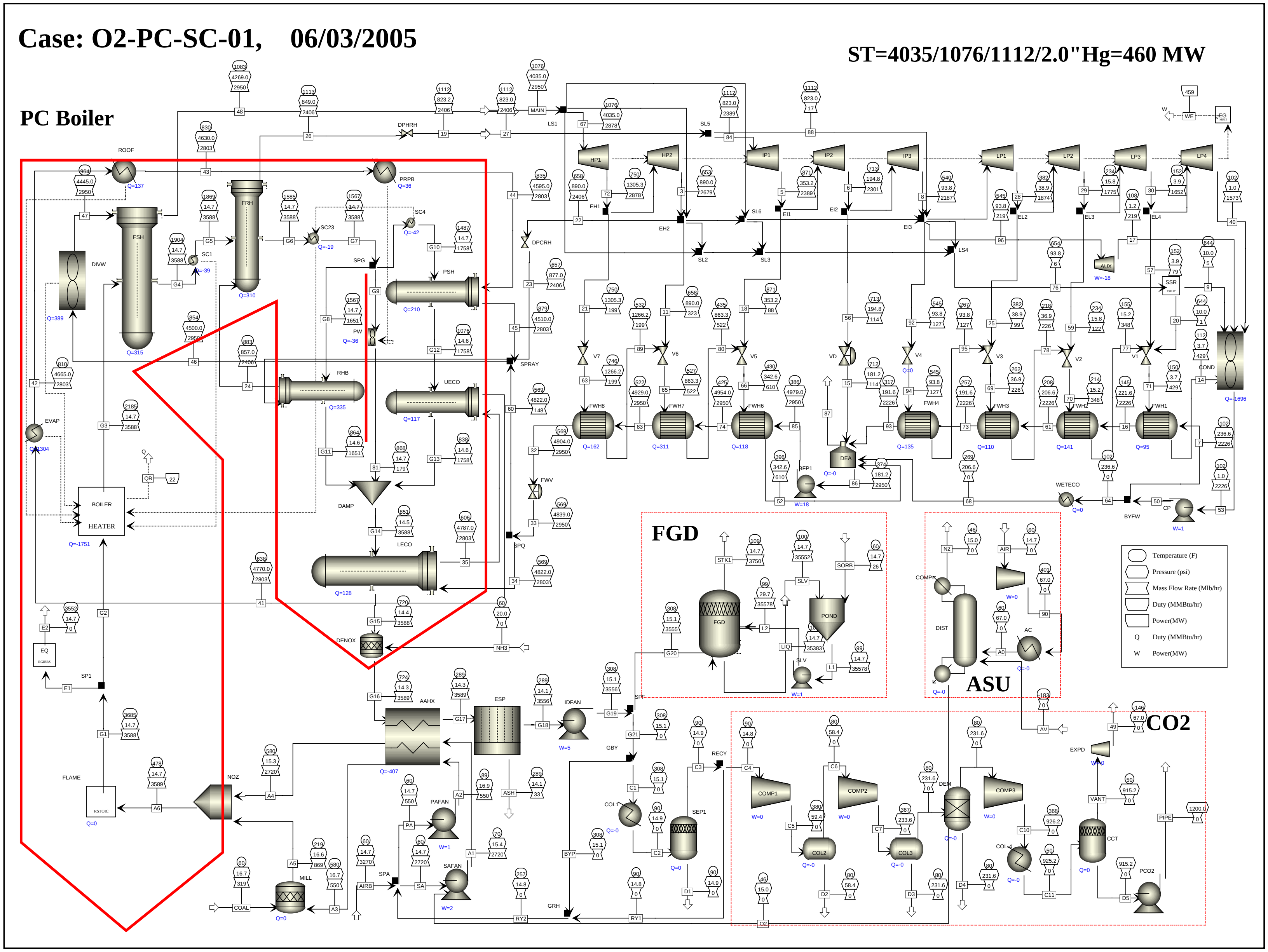
The temperature of the flue gas exhausted to the stack is 289ºF. The flue gas exiting the boiler contains 3.0%, vol., wet O_2 (18% excess air) and contains 739 klb/hr (1.72 lb/kWh) of CO_2. This 3.0% O_2 level is kept constant for all of the O_2-fired cases. A SCR is applied to control NOx with NH_3/NOx=1.0, while an FGD is used to control SOx by lime solution with Ca/S=1.05, L/G=10, and 85% excess air for aeration.

The breakdown of auxiliary power for case 1 is listed in Table 2. Most of these power consumptions were simulated directly by the Aspen module. Some required user Fortran for those processes lacking Aspen modules, such as solids handling. The power consumption was based on stream flows and design data.

Fan power consumption was simulated based on the pressure drops from both air side and gas side. The total auxiliary power consumption, including FGD, for case-1 is approximately 9.7% of the gross power, while it was about 9.2% for a subcritical case.

Because of high temperature and high CO_2 concentration, the CO slip becomes high, and the actual flame temperature is lower than that of adiabatic combustion temperature. To represent this effect on boiler design, estimation of equilibrium flame temperature was modeled, where the equilibrium CO concentration in flue gas was calculated.

Figure 1 - Reference Case of Air-fired Supercritical PC



4.3 Oxygen-Fired PC Integrated with Cryogenic ASU

4.3.1 Boiler Plant Modifications

The oxygen-based (or oxygen-fired) plant model was simulated as reported in Task 1. It contains essentially all the components in the conventional plant model. In addition, it includes a cryogenic air separation unit (ASU) and a flue gas cooler. In the O_2-fired plant, the FGD is not needed because the SO_2 is acid gas similar to CO_2 and can thus be sent to pipeline together with the CO_2. A substantial portion of the SOx will be removed as the flue gas is cooled down in the CO_2 cooling and compression equipment.

The steam side components remain very similar to the air-fired case with only some changes in heat bundle duties in the heat recover area (HRA).

In O_2-fired cases, flue gas is recycled to control the flame temperature inside the PC-fired boiler to minimize NOx formation, minimize ash slagging in the furnace combustion zone, and avoid the application of exotic materials.

Before the flue gas is separated into a recycled and effluent stream (to the pipeline), it is cooled to about 90ºF. Since this is below the acid/moisture dew point, a heat exchanger containing acid-resistant materials must be used. The recycled gas is then reheated, before the forced draft (FD) fan, by mixing it with a by-passed hot gas to avoid reaching the dew point. After the O_2 from ASU plant is mixed with recycled flue gas, it is heated by the flue gas exiting the boiler in a gas-to-gas heat exchanger, which acts as a recuperator to improve cycle efficiency and reduce fan power requirements.

It is assumed in this study that there is no tramp air ingress through the sealed boiler.

4.3.2 Cryogenic Air Separation Unit

A commercially available large-scale cryogenic air separation technique was used for O_2-fired PC base case. A traditional cryogenic ASU plant was simplified in the simulation to include the power consumption, but without details of distillation columns and cold heat exchangers. The Aspen model does not include the air purifier, which removes moisture, hydrocarbons, CO_2, and NOx in an adsorber and is located between the cold box and air compressor. Although the separated N_2/Ar gases could potentially be sold as byproducts, no economic credit for this is taken in this study. No heat recovery from the ASU air compressor inter-stage coolers is included, because recovery of this low-grade heat is inefficient.

For the O_2 purity of 99.5% used in the Task-1 study, a power consumption of 24.5 kWh/klb$_{air}$ was applied. For a 460 MW steam turbine generation, the ASU plant consumes about 70 MW, or 15% of generated power, which is a large penalty for CO_2 removal.

4.3.3 CO_2 Compression Unit

The flue gas effluent stream (mainly CO_2) has to be compressed to the high pipeline pressure of 1200 to 3000 psia depending upon end user specification. The CO_2 sequestration equipment is added to the system and the effluent is conservatively compressed to 3000 psia. The dominant moisture in flue gas is condensed out first during flue gas cooling before the first stage compression. The condensed water contains acid gases and has to be treated before recycle or discharge.

The flue gas composition after cooling (but before the first stage CO_2 compressor) from this base case (case-6) is:

CO_2	O_2	N_2+Ar	SOx	H_2O
90.4	3.3	1.3	1.1	3.6

In the literature, O_2 as low as 1.3% was used. Reducing O_2 content, such as from 3.0% to 2.0% by reducing excess air, is helpful in reducing CO_2 compression power, but it is judged that and oxygen content of approximately 3.0% is required for good combustion efficiency. Since CO_2 and SO_X are both acid gases, they can combine with moisture to form acid, which can cause corrosion in the CO_2 pipeline. Therefore, after the 2nd stage, a chemical method of active dehydration with TEG (Triethyleneglycol), regarding hydrate formation and corrosion, has been applied to remove the rest of moisture out to a very low level (less than 50 ppm), where the TEG can be regenerated by heating. In the model, the TEG dehydration was simulated, but the TEG itself was not simulated.

A four-stage flue gas CO_2 compression with inter-stage cooling was applied. To reduce power, an equal compression pressure ratio of approximately 4.0 was applied.

4.3.4 O_2-fired PC with Cryogenic ASU (case-6)

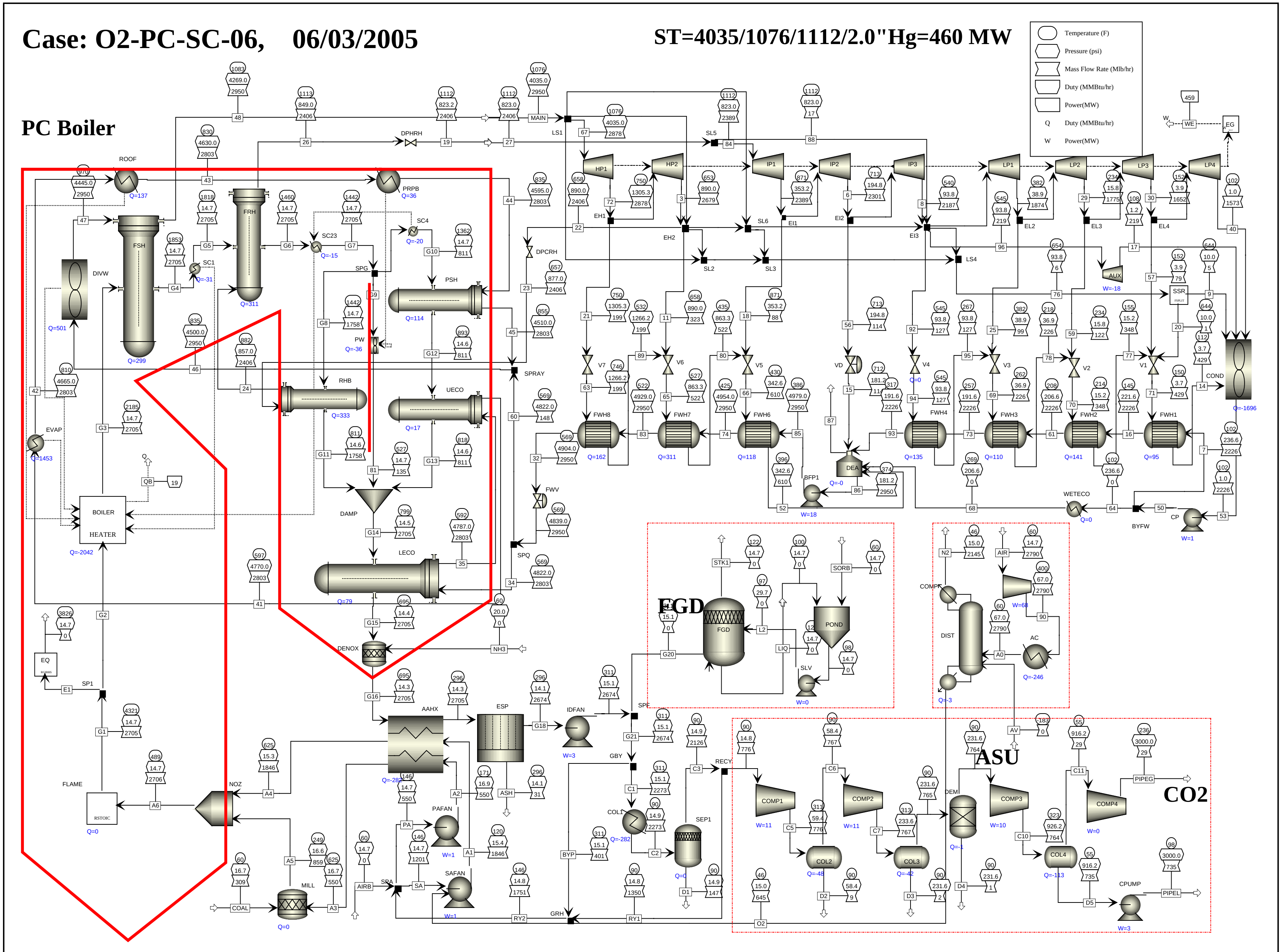
Figure 2 presents the base case system model. The boiler equilibrium temperature of 3800ºF was kept at approximately the same as that of the reference air case by adjusting the recycle gas flow. Both fuel and air feed rates were iterated to meet the heat duty and 3.0%v exit O_2 level (this corresponds to a boiler excess of 10.5% and a net excess of 3.3%). Compared to reference air-fired case, the case-6 air flow rate is reduced by 15% (from 3270 klb/hr to 2790 klb/hr) and O_2 concentration to the boiler is 33.8% (compared to 20.7% for the air-fired case) yielding an increased combustion efficiency and a lower UBC. The relation of UBC and O_2% has been included in the modeling as reported in Task 1. The extra cooling duties for the CO_2 stream were added into cooling tower calculation for auxiliary power.

The oxidant from the ASU was 99.5% O_2 pure. ASU and CO_2 plant power consumptions were 67.7 and 34.2 MW, respectively, resulting in a total auxiliary power of 138.5 MW, or about 29.1% of gross power. All CO_2 generated from fuel combustion was 100% removed with a final CO_2 purity of 93.8%v. The CO_2 removal efficiency drop was 7.5% in points (39.46% to 31.94%), and the power penalty was 114 kWh/klbCO_2.

Oxygen-firing versus air-firing reduced the flue gas volume flow from 179 to 96 MMft^3/hr, or to 54%, due to a high gas molecular weight of CO_2 instead of N_2. The gas mass flow was reduced from 3555 to 2675 klb/hr, or 75% in comparison to the air-fired case. The recycled gas flow was 1751 klb/hr, with a temperature of 146ºF, preheated by a bypass flue gas stream as shown by Figure 2. Compared to the air-fired case, the coal feed rate was reduced from 319 to 309 klb/hr as the result of low UBC, and low excess air.

Because of high CO_2 content in flue gas at high flame temperature, part of CO_2 dissociated and more CO slipped, producing an equilibrium flame temperature of 3830ºF, compared to the adiabatic temperature of 4321ºF. The actual flame temperature will be even lower than the equilibrium flame temperature when over-fired gas is applied.

Figure 2 - Base case of O_2-fired supercritical PC



4.4 O_2-Fired PC Integrated with OITM

Advanced oxygen separation through high temperature membranes such as OITM is currently under development [2, 4, 5, 6]. Oxygen ion transport membranes have the potential to provide a major reduction in oxygen separation capital and energy consumption.

Oxygen ion transport is driven by the difference in oxygen partial pressure across a membrane. Oxygen atoms adsorb on the cathode (high oxygen partial pressure side of the membrane) and dissociate into ions as they pick electrons. These ions travel from cathode to anode (the low oxygen partial pressure) by jumping through lattice sites and vacancies until they reach the anode side of the membrane. On the anode side, the oxygen ions yield their electrons to become atoms/molecules, which are then desorbed into the gas phase. Electrons from the anode side are carried through the membrane to the cathode side to complete the circuit. The rate of oxygen transport through such membranes is temperature sensitive, and can be very fast at high temperatures. These membranes have infinite selectivity for oxygen over other gases, because only oxygen ions can occupy the lattice positions. A typical schematic of the oxygen ion transport membrane process is presented in Figure 3.

The OITM process and design data are based on Reference 6. A schematic of the OITM process incorporated within the O_2-PC plant is shown in Figure 4. To integrate the OITM into the O_2-fired PC, air is pressurized by a compressor, and then heated to a high temperature. High air pressure provides a high oxygen partial pressure on the airside of the OITM to reduce the size and cost of the OITM. Air enters the compressor at 60ºF and 14.7 psia and is compressed at 85% efficiency to 215 psia and 743ºF. The compressed air is heated to 1652ºF by a tubular air heater inside the PC furnace. This hot pressurized air is fed to the OITM, where about 85% of its O_2 is separated through the membrane to an exit pressure of 16 psia, and the rest of the vitiated air or O_2-depleted air is sent to a gas expander to generate power at 86% turbine efficiency. Power generated from the expander is used to drive the air compressor. Since the power generated in the gas expander is greater than the air compressor power the OITM system produces a net power output.

The separated O_2 from the membrane is carried by a heated sweep gas. The use of a sweep gas reduces the oxygen partial pressure on the low-pressure side of the membrane and consequently reduces the size and cost of the OITM. A recuperator is applied between inlet and outlet sweep gas flows to reduce the heat requirement. Heat transfer from the sweep gas to the membrane to the air is neglected as it is expected to be relatively small (it was also neglected in Reference 6). The mixture of sweep gas and O_2 is then injected into the furnace. In this design, the OITM does not consume electrical power; instead, it absorbs heat, generates power, and separates O_2 from air. Therefore it is expected to reduce ASU operating and capital costs.

From a heat and mass balance point of view, if there is no heat transfer between the air and sweep gas, the model can be simplified by directly mixing the separated O_2 with the sweep gas. Thus, to simplify the model, the recuperator and OITM are combined into a single module in the Aspen model. The operational details of the OITM, such as the O_2 flux through the membrane as function of OITM temperature and pressure, were not included in the Aspen model. These details are necessary only for OITM size and cost estimations. Moreover, the Aspen model does not directly model the recuperator, but its design and configuration is required for the economic analysis.

Without performing a detailed economic study, it is difficult to determine the optimum OITM operating pressure. However, Reference 7 specifies that for an economic design the ratio of oxygen partial pressures of the feed gas (air) to the permeate stream (sweep gas) should be approximately 7. The base case OITM design has an oxygen partial pressure ratio of 4.9 at the air inlet and 10.2 at the air outlet yielding an approximate average ratio of 7.5. Reference 7 specifies that an 85% O_2 recovery and an operating temperature of 1652ºF is within the operating range of the OITM. The effect of O_2 recovery and OITM operating pressure on system performance and OITM design is examined in Section 4.4.3.1 and Section 4.4.3.2, respectively.

Several cases incorporating an OITM ASU into the O_2-PC have been simulated to evaluate the effect of different conceptual designs and operating conditions on power plant performance. For the OITM ASU cases, the specific power penalty for the CO_2 removal cannot be calculated by difference directly, because extra power is produced through increased coal-firing. In some cases, the extra power by increased coal-firing can be greater than the power required for CO_2 removal. Therefore, the CO_2 removal power penalty is defined indirectly as

$$\text{kWh/lbCO}_2 = (\text{eff drop})*(\text{power/efficiency})_{\text{air}}\ /\text{lbCO}_2 \qquad (1)$$

Equation (1) is a way to compare the penalty of CO_2 removal for different systems without cost estimation, especially for a complex system, which fires more when CO_2 removal is involved, such as the OITM O_2-PC.

Figure 3 - A Typical Oxygen Ion Transport Membrane Schematic [4]

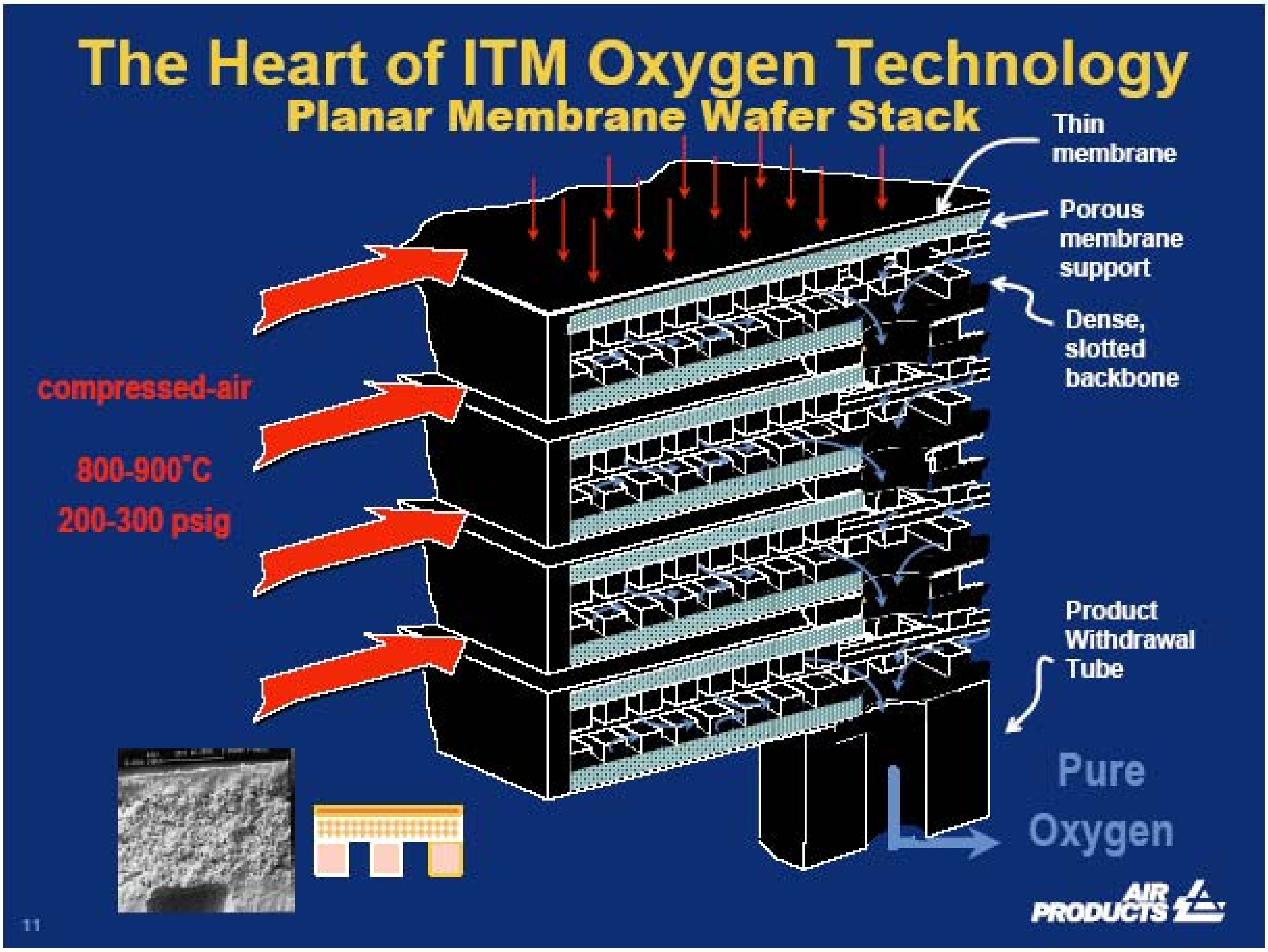
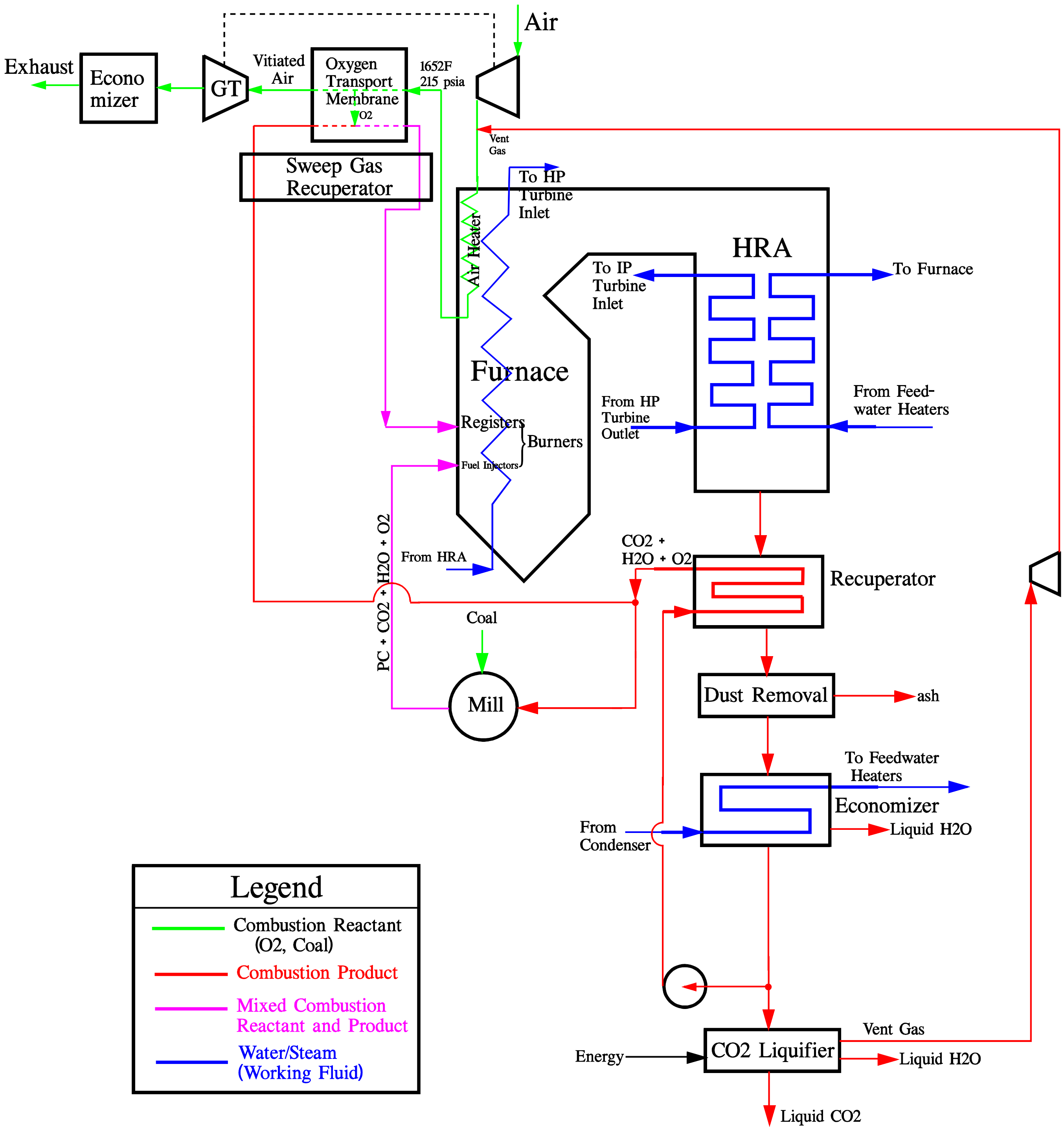


Figure 4 – Integration of Oxygen Ion Transport Membrane into O2-PC



4.4.1 Process Gains

The inclusion of a gas turbine expander in the O_2-PC power plant utilizing the OITM increases the overall system efficiency by allowing work to be done in the gas turbine at a higher temperature than can be achieved by the steam from the boiler. This principle has been applied by FW in its 1st generation Pressurized Fluidized Bed Combustor (PFBC) design. A further enhancement is the 2nd generation PFBC design (similar to IGCC) in which the turbine entrance temperature is increased by syngas combustion. This concept was also applied in the FW High Performance Power System (HIPPS) design, which includes an in-furnace high temperature air heater similar to the one required for the O2-PC OITM concept.

4.4.2 Base Case (case-11)

Figure 4 shows a schematic of the OITM ASU integrated with the O_2-fired PC power plant. The OITM system includes a sweep gas system, and an air supply system. The pressurized vent gas from the CO_2 plant is recycled back to the air separation unit, where it is mixed with the compressed air. In this way the rich O_2 in the vent gas can be recovered, and the air to compressor can be reduced. Therefore this vent gas recycling increases system efficiency and reduces the operating cost.

Figure 5 is a process flow diagram generated by Aspen Plus for the OITM ASU integration. Air is compressed to about 200 psia by a compressor, and then heated within the boiler to about 1650ºF. This hot pressurized air is fed to OITM, where about 85% of its O_2 is separated through a membrane, and the rest of vitiated air is sent to an expander. The separated O_2 from the membrane is carried by a heated recycled flue gas after gas-to-gas heat exchange. A recuperator is applied between inlet and exit sweep gas flows (not shown in Figure 5 since it lumped inside the OITM module). The mixture of sweep gas and O_2 is fed directly to the boiler. The exhaust gas from the expander passes through an economizer to release its heat for feedwater heating. Power generated from the expander is used to drive the air compressor.

The O_2 obtained from the OITM is swept with recycle flue gas. Since the compressed air is heated to 1650ºF inside the boiler, a special heat exchanger and boiler design has to be used for the OITM application (this will be described in the Task 3 report). The boiler air heater duty is 974 MMBtu/hr for Case-11. The coal feed rate is 377 klb/hr as compared with 319 klb/hr for the air-fired case-1, and 309 klb/hr for the O_2-fired (cryogenic ASU) case-6. The corresponding flue gas flow increased to 3497 klb/hr, nearly approaching the air-fired flow of 3552 klb/hr. The boiler O_2 concentration is 31%v, which is nearly the same as case-6.

Because of increased flue gas flow created by greater coal-firing, more heat is carried to the boiler HRA in case-11 than in case-6. As result, distribution of heat duty is shifts as for example, the heat duty of the division wall reduces from 501 MMBtu/hr in case-6 to 396 MMBtu/hr in case-11, while the sum of the primary superheater and upper economizer duties increases from 131 MMBtu/hr in case-6 to 411 MMBtu/hr in case-11. Consequently, the inlet furnace feedwater temperature increases from 597ºF in case-6 to 666ºF in case-11. The total furnace duty for case-11 increases because of the 974 MMBtu/hr air heater duty, although the heat to waterwalls is reduced from 2042 to 1583 MMBtu/hr.

Because more low-grade heat is released from the gas turbine (GT) exhaust and from boiler, all low-pressure feedwater heaters are shut off. The low-pressure feedwater heating is provided in parallel by the GT economizer and a HRA flue gas exhaust economizer (in parallel to flue gas heat recuperator), as well as by part of the compressor inter-stage coolers.

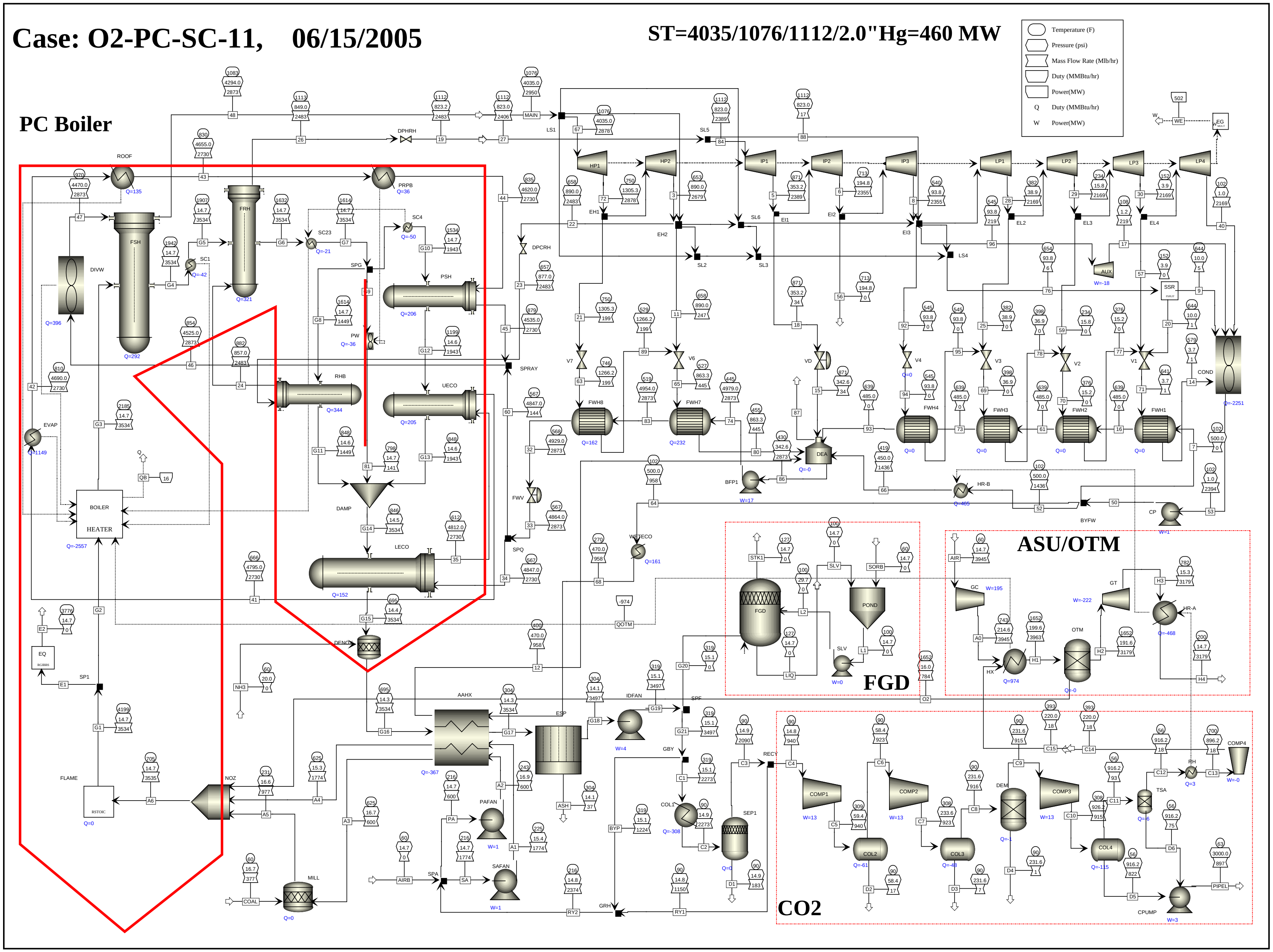
The OITM O_2-PC incorporates the recycling of CO_2 compression system vent gas back to ASU. The high-pressure vent gas from the CO_2 plant is heated with GT exhaust gas before it is expanded for power recovery. After the expansion to the OITM operation pressure, the O_2-rich vent gas is mixed with compressed air and sent to OITM. Consequently, the compressor air flow and power is reduced. The emission control equipment can treat the vent gas prior to the OITM to require smaller equipment sizes due to the high pressure of the vent gas.

The replacement of the cryogenic ASU with the OITM greatly reduces the efficiency loss penalty. This is caused by lower equivalent ASU power, better system integration, increased power from the OITM GT, and more low-grade heat recovery from cooling. As shown in Table 3 the OITM reduces the efficiency drop by 52% and the CO_2 removal power penalty by 61%.

Table 3 – Comparison of O2PC with Cryogenic and OITM ASU

	Air-fired	O2-fired	O2-fired
ASU	-	cryogenic	OITM
Main steam flow, klb/hr	2950	2950	2950
Coal flow, klb/hr	319	309.4	377
Net power, MW	430	338	462
Net efficiency, %	39.5	31.9	35.8
ASU Power, MW	0	67.7	-26.7
CO2 Compression Power, MW	0	34.2	41.3
CO2 removal flow, klb/hr	0	720	874
Efficiency penalty, % points	0	7.5	3.6
CO2 removal penalty, kWh/klbCO2	0	114	45

Figure 5 - O2-PC with OITM



4.4.3 Parametric Studies

Several parametric cases were run as described in the following sections. The results are summarized in Table 4.

Table 4 – Parametric Case Summary

case		01	06	09	11	12	13	14	15	16	17	18
Air separation method		None	Cryo	Cryo	OITM	OITM	OITM	OITM	OITM	OITM	OITM	OITM
Waste heat economizer		no	no	yes	yes	yes	yes	yes	yes	yes	yes	yes
HRA Arrangement		parallel	parallel	parallel	parallel	parallel	parallel	parallel	parallel	parallel	parallel	paralle
CO2 Condensation		-	yes	yes	yes	yes	yes	yes	yes	yes	yes	yes
Vent Gas Recycle		-	no	no	yes	yes	yes	yes	yes	yes	yes	yes
Coal flow	klb/hr	319.0	309.4	308.0	377.7	376.0	403.8	377.7	377.7	377.7	389.4	375.4
Oxidant flow	klb/hr	3270	645	644	784	780	838	784	784	784	808	781
Air flow	klb/hr	3270	2790	2784	3945	3685	4945	4284	3810	3875	4067	3930
O2 purity	%	-	99.5	99.5	100	100	100	100	100	100	100	100
Recycle gas flow	klb/hr	0	1751	1166	2374	2356	2523	2374	2374	2374	2392	1525
	%	0.0	65.5	55.9	67.9	67.8	67.8	67.9	67.9	67.9	67.4	57.7
Recycle gas temperature	F	-	146	260	216	217	223	214	217	216	220	249
Boiler Inlet O2	%, v	20.7	33.8	41.1	31	31.1	31.1	31	31	31	31.4	39.3
Boiler Outlet O2	%, v	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0
Boiler outlet flue gas flow	MM cf/hr	179	96	77	139	138	150	139	139	139	142	100
	klb/hr	3555	2675	2087	3497	3474	3723	3497	3497	3497	3550	2644
Adiabatic Temperature	F	3685	4321	5104	4196	4202	4208	4196	4196	4196	4243	5120
Equilibrium Temperature	F	3552	3830	4161	3770	3773	3775	3770	3771	3771	3792	4169
Gas Temp. to FSH	F	2185	2185	2185	2185	2185	2185	2185	2185	2185	2185	2185
Water temp. to evap.	F	638	597	596	666	664	690	666	666	666	672	599
UBC	%	1.00	0.30	0.16	0.38	0.38	0.38	0.38	0.38	0.38	0.38	0.16
Boiler Efficiency	%	88.16	90.90	91.31	-	-	-	-	-	-	-	-
Pipeline pressure	psia	0	3000	3000	3000	3000	3000	3000	3000	3000	3000	3000
Generated CO2 flow	klb/hr	739	720	718	874	869	934	874	874	874	900	866
CO2 purity	%,v wet	14.0	77.5	71.3	75	74.9	74.9	75	75	75	74.8	70.8
Removed CO2 flow	klb/hr	0	720	718	874	869	934	874	874	874	900	866
CO2 removal efficiency	%	0	100	100	100	100	100	100	100	100	100	100
CO2 purity	%	-	93.7	93.6	96.9	96.9	96.9	96.9	96.9	96.9	96.9	96.9
Gross Power	MW	476.2	476.2	486.4	519.7	515.8	520.3	519.3	520.0	519.8	521.8	519.5
Auxiliary Power	MW	46.0	36.6	38.1	43.0	42.8	44.0	43.1	43.0	43.0	43.4	41.6
ASU power	MW	0.0	67.7	67.6	-26.7	-22.0	-42.2	-24.3	-27.3	-27.1	-35.9	-26.6
CO2 compression power	MW	0.0	34.2	34.2	41.3	41.1	44.1	41.3	41.4	41.4	42.6	41.2
Net Power	MW	430.2	337.7	346.6	462.1	453.9	474.4	459.2	462.9	462.5	471.7	463.3
Net Efficiency	%	39.46	31.94	32.93	35.80	35.33	34.38	35.58	35.87	35.83	35.45	36.11
Efficiency Drop	% pts.	-	7.5	6.5	3.7	4.1	5.1	3.9	3.6	3.6	4.0	3.3
CO2 removal energy	kWh/klbCO2	-	114	99	46	52	59	48	45	45	49	42

4.4.3.1 Effect of O_2 Recovery Efficiency (case 11 to 13)

As described in Section 4.4.1, shifting more duty to the GT will increase system efficiency. Increasing OITM O_2 recovery efficiency reduces the required air flow rate and decreases the GT mass flow and power generated.

Case-12 is the case where the O_2 recovery is increased from 85% to 90.5%. As a result of high O_2 recovery, less air is required to be fed to the system, and so less power is generated in the GT. This leads to a reduction of system efficiency from 35.8% to 35.3%.

In Case-11 and Case-12 the low-grade heat released from GT exhaust, from flue gas cooling before CO_2 compression, and from CO_2 compressor inter-stage cooling is recovered to heat the low pressure feedwater. In Case-13, when the O_2 recovery is reduced from 85% to 73%, more air has to be fed to OITM (4945 vs. 3945 klb/hr) and so more heat is required by the OITM. The system fires more coal (404 vs. 378 klb/hr) and as result, the system generates more low-grade heat than can be recovered, which results in an increase of GT exhaust temperature from 200ºF 330ºF. This reduces the system efficiency even with increased extra power from the GT.

It is clear that the two opposing effects, more GT power and more low grade heat from increased air flow, form a system with an optimum performance for a given air side pressure as shown by Figure 6, where the system efficiency is maximum and the CO_2 removal power penalty is minimum when the O_2 recovery is approximately 86%. Note that recovering additional low-grade heat as through the use of a high-pressure economizer, will shift the optimum O_2 recovery efficiency. As more of the low grade heat is recovered by the use of more complex heat integration schemes or by co-generation heat export, then the optimum O_2 recovery efficiency is reduced and the maximum system efficiency is increased. Such a reduction in O_2 recovery efficiency reduces the size of the OITM because of increased logarithm mean pressure difference (LMPD).

Table 5 shows a comparison of different cases under the same airside pressure, including a case published by Alstom [6]. It is obvious that the higher is the O_2 recovery efficiency, the lower is the LMPD, and the larger is the OITM size.

Figure 6 – Effect of OITM O_2 Recovery Efficiency on Efficiency

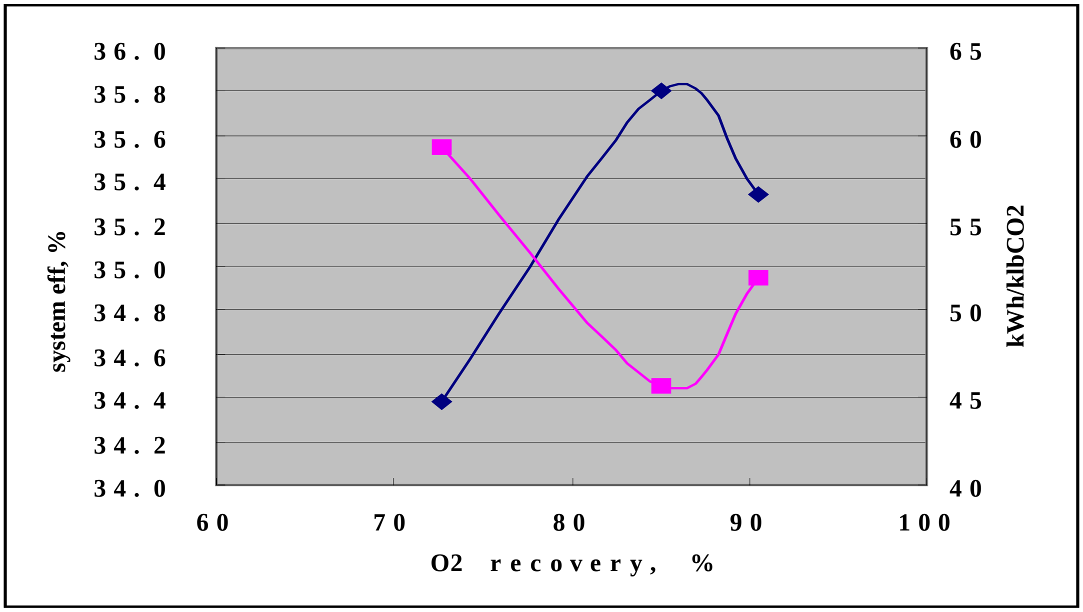


Table 5 - Effect of OITM O_2 Recovery Efficiency on LMPD

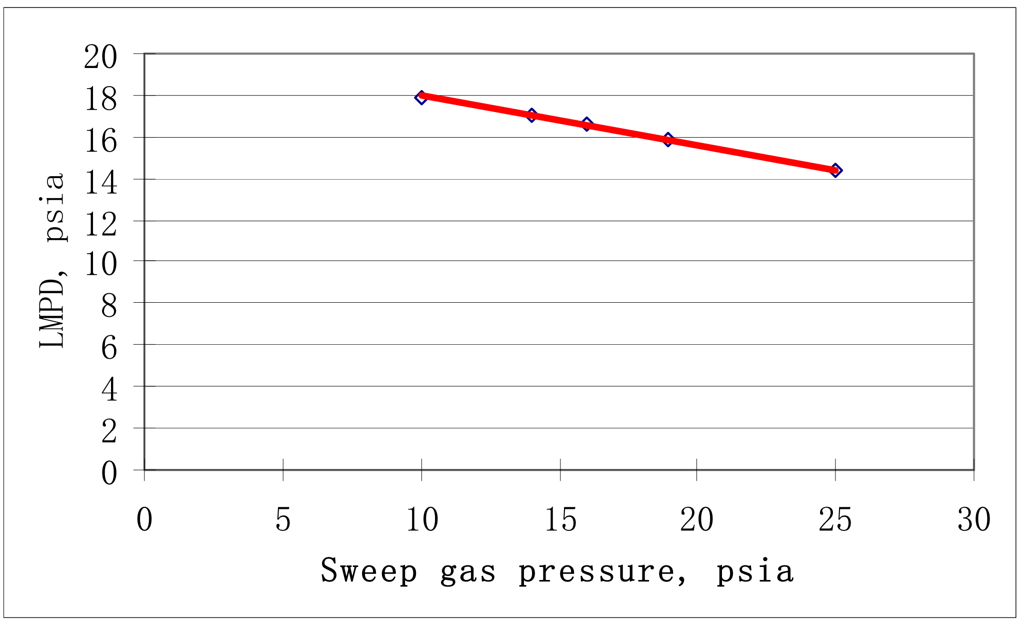
Case	Alstom	Case-11	Case-12	Case-13
O2 recovery, %	85	85	90	73
O2 to boiler, %	70	31	31	31
LMPD	14.7	16.5	13.9	20.9

4.4.3.2 Effect of LMPD Across the Membrane (cases 14-16)

The design of the OITM relies on the O_2 partial pressure difference across the OITM membrane. Similar to the LMTD used in a heat transfer process, a LMPD can be applied for a mass transfer process. The LMPD can be used to compare the performance of different options.

An increase in the O_2 level to the boiler increases the sweep gas outlet O_2 partial pressure resulting in a reduced LMPD as shown in Table 5 (compare Alstom to case-11). Similarly an increase in sweep gas total pressure reduces the LMPD as shown in Figure 7. The effect is fairly small because of the limited operating pressure variation of the sweep gas.

Figure 7 – Effect of Sweep Gas Pressure on LMPD



The air side pressure directly affects the OITM performance. Increasing the air side pressure (OITM operating pressure) will:

(1) increase LMPD, and so reduce OITM size
(2) increase compressor discharge temperature (CDT), and require less heat from the boiler
(3) reduce turbine exhaust temperature (TET), and so release less low grade heat and reduce the optimum O_2 recovery efficiency
(4) increase equipment thickness

In Case-14 the OITM operating pressure is raised from 200 psia to 250 psia with the same coal feed rate and furnace air heater duty. This requires the compressor discharge pressure to be increased from 214 to 265 psia, which raises the compressor discharge temperature. Thus, for the same operating temperature the OITM needs less heat per unit mass of air, and so the air to the OITM was increased to balance the heat released from the boiler. Because the amount of the O_2 required is fixed, the O_2 recovery is adjusted for the increased air flow through the OITM. Figure 8 shows that raising the air side pressure from 200 psia to 250 psia results in a small reduction in system efficiency (from 35.8 to 35.6%) and an attendant small increase in the CO_2 removal specific power penalty (from 46 to 48 kWh/klbCO_2). The system efficiency decreases with increased OITM pressure because of less heat carried to the OITM per unit air. However, due to the increased air flow to the OITM, the LMPD is increased from 16.5 to 24.9 psia, which results in a decrease in OITM size of 34%.

Opposite to the case-14, case-15 and 16 were run with reduced OITM pressures of 180 psia and 190 psia, respectively. As shown in Figure 8, lower OITM operating pressure (for a given operation temperature) increases system efficiency because more heat is transferred from the boiler to the OITM cycle per unit mass of air. However, as the OTM pressure is reduced the LMPD decreases requiring a larger OITM size.

Figure 8 – Effect of OITM Pressure on System Efficiency and LMPD: Variable O2 Recovery

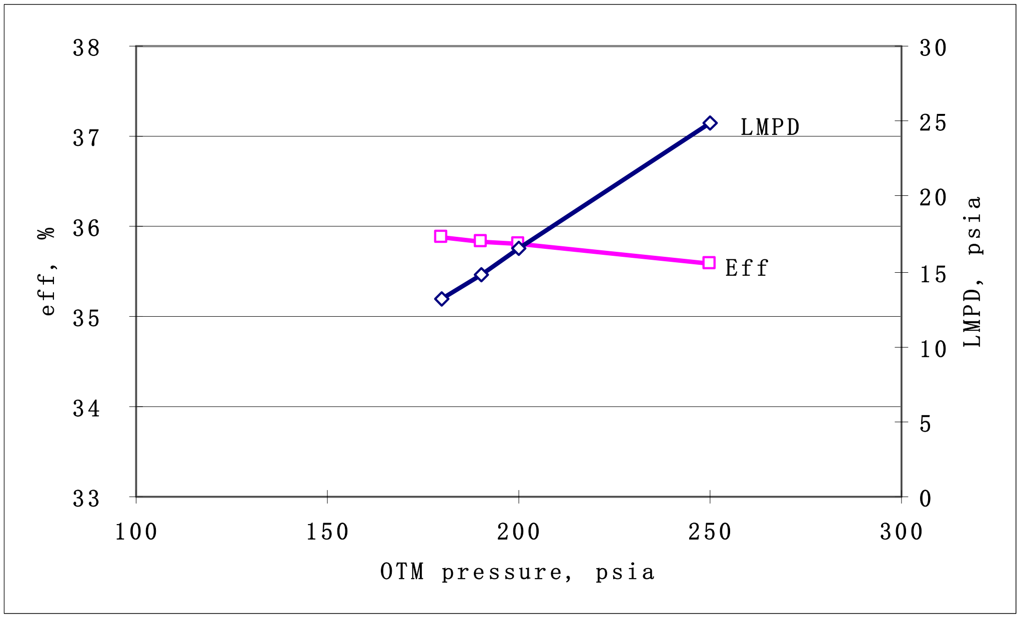
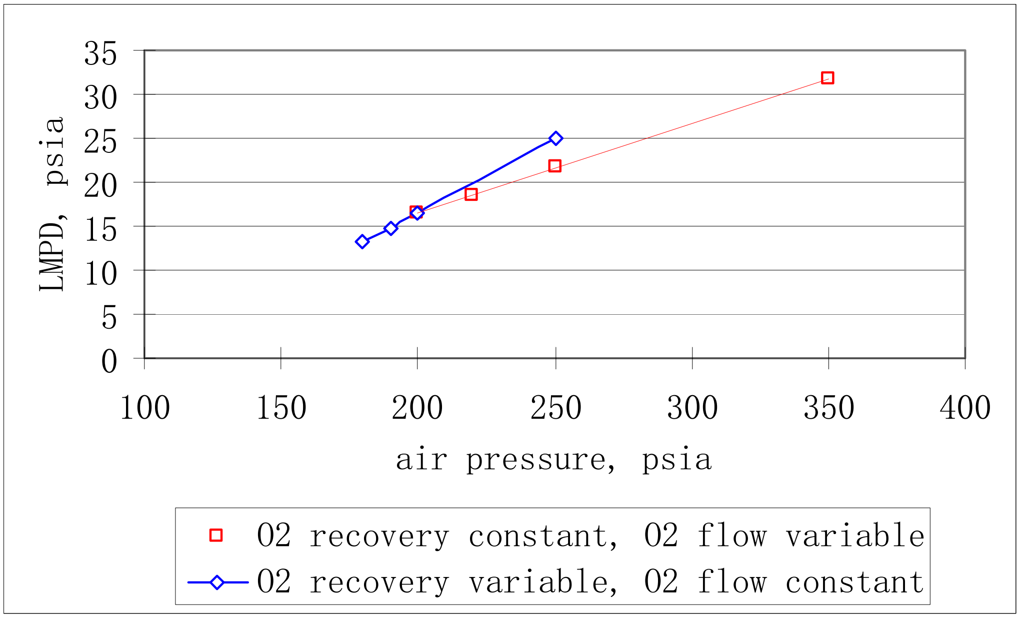


Figure 9 shows the effect of OITM pressure on LMPD when the O_2 recovery efficiency is constant and O_2 flow rate is variable. This is compared in Figure 9 with the effect of OITM on LMPD when the O_2 flow rate is constant and the O_2 recovery efficiency is variable (as in Figure 8).

The selection of the OITM operating pressure is a trade-off between the cost of OITM and the system efficiency. If the OITM cost is not too high in future commercial application, the optimum OITM operating pressure will be relatively low. Furthermore, a higher OITM operation temperature will be better for the system efficiency. However, the magnitude of this temperature is constrained by material limitations.

Figure 9 – Effect of OITM Pressure on System Efficiency and LMPD: Constant and Variable O2 Recovery



4.4.3.3 Effect Of Compressor Discharge Temperature (case 17)

Another way to boost the system efficiency is to transfer more heat to the gas turbine by increasing the temperature difference between the compressor discharge temperature (CDT) and the turbine inlet temperature (TIT). Increasing the TIT results in an increase in OITM operating temperature, which may be restricted due to material limitations. Without increasing the pressure, the increase of TIT will increase turbine exhaust temperature, which results in more low-grade heat to be recovered from HRSG. Another approach is to reduce CDT by the use of a compressor with inter-stage cooling. This will reduce compressor power, and let more heat be transferred from the boiler to the turbine, but it releases more low-grade heat from inter-stage cooling. If this heat cannot be recovered, system efficiency would be reduced. As discussed before, for the present configuration and integration, there was no margin to recover more low-grade heat. Therefore the inter-stage cooler will be applicable only for co-generation of heat and power, where low-grade heat could be recovered by low-pressure steam export.

Case-17 employs an alternative method to reduce compressor power by inter-stage water quench to avoid the need for low-grade heat recovery. This quench (20 klb/hr of water) reduced the CDT from 743 to 698ºF, and therefore more heat flowed from the boiler to the gas turbine. As a result, the coal to boiler increased from 377.7 to 389.4 klb/hr, and the corresponding air to the OITM increased from 3945 to 4067 klb/hr (for the same OITM O_2 recovery efficiency). The power from the GT increased from 26.7 to 35.9 MW, and the net power increased from 462 to 472 MW. However, the system net efficiency was reduced from 35.80 to 35.45%, and the corresponding net penalty for CO_2 removal increased from 46 to 49 kWh/klbCO_2 because of efficiency loss. The great benefit from this option was the 10 MWe net power gain. Note also that the furnace air heat duty increased from 974 to 1060 MMBtu/hr.

4.4.3.4 Effect of Furnace Flame Temperature (case 18)

Increased furnace flame temperature increases heat transfer, especially for radiant transfer, and so it will reduce the furnace size for both for the waterwalls and air heater. Similar to the effect of excess air in the boiler, higher flame temperature slightly increases system efficiency because of less flue gas flow out of the system.

Higher flame temperature cases have been evaluated for the cryogenic ASU O_2-PC, as reported in Task 1 of this project. For the cryogenic ASU O_2-PC raising the equilibrium temperature from 3830ºF (case 6) to 4182ºF (case 7) increased system efficiency about 0.15% in point, and reduced specific power penalty for CO_2 removal from 114 to 112 kWh/klbCO_2 as shown in Table 6.

Case 18 was generated from case 11 by raising the flame temperature to nearly the same flame temperature as case 7. Table 6 shows that for the OITM ASU O_2-PC raising the equilibrium temperature to 4169ºF increases system efficiency by about 0.3% in point, and reduces the CO_2 removal specific power penalty from 46 to 42 kWh/klbCO_2. It is clear that increased flame temperature can reduce equipment size and slightly improve system efficiency, but could be limited by material cost in the furnace.

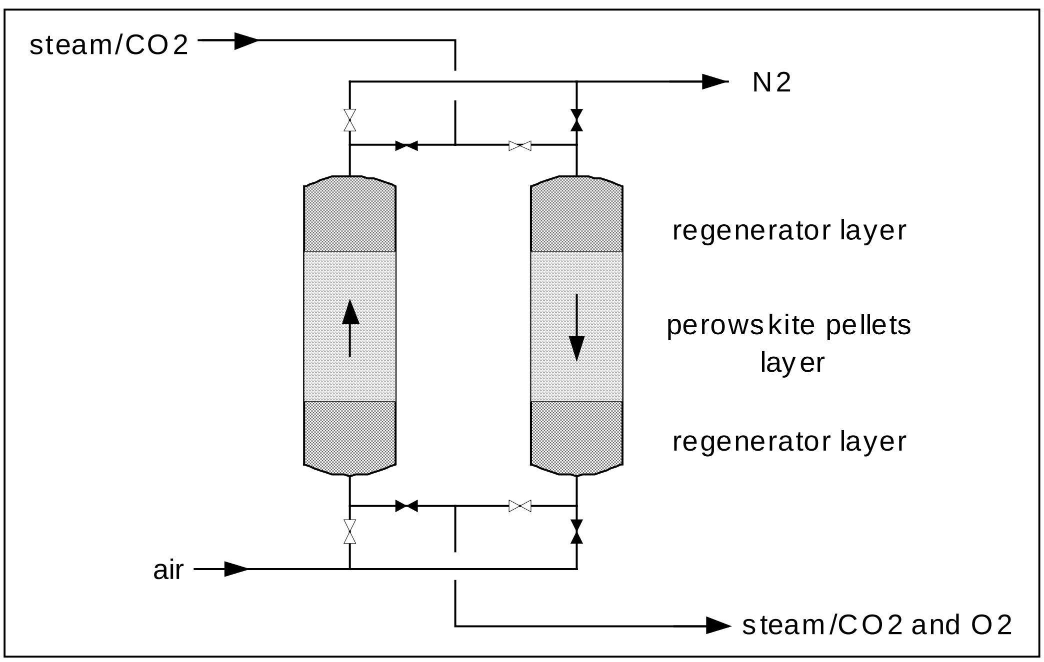
Table 6 – Effect of Flame Temperature on Performance

Case	6	7	11	18
Adiabatic Flame T, F	4321	5178	4196	5120
Equilibrium flame T, F	3830	4182	3770	4169
Recycle flue gas T, F	146	163	216	249
system eff, %	31.94	32.07	35.80	36.11
kWh/klbCO2	114	112	46	42

4.5 O_2-Fired PC Integrated with CAR

The ceramic auto-thermal recovery (CAR) process [5] is based on sorption and storage of oxygen in a fixed bed containing ionic and electronic conductor materials operated at high temperature and increased pressure. The stored oxygen is then released by pressure reduction using sweeping gas, such as recycled flue gas, or steam extracted from low-pressure section of steam turbine as shown in Figure 10. The continuous operation is obtained by employing multiple beds in a cyclic way, which is similar to the Pressure Swing Absorption (PSA) process. A large vessel is provided to provide a five second buffer time to smooth out any fluctuations in either flow and/or composition caused by a batch adsorption-desorption operation cycle. An important feature of the CAR process is in that it can be tailored to produce low-pressure oxygen at the concentration required for O_2-fired combustion by using recycled cleaned flue gas as a sweep gas. The CAR process is based on conventional sorbent bed adsorption that is easy to fabricate and readily available as claimed. The scaling up for such a process is similar to the PSA process, and has fewer challenges than the OITM technology.

Figure 10 - CAR Process Schematic



In general during adsorption, heat is released and system temperature is raised. The CAR adsorption process has to be operated at high temperature, therefore the air is preheated to a certain temperature before being fed to the adsorption bed. The heat generated during adsorption can be recuperated to heat up the fresh air to reduce heat loading. For this purpose, a recuperator is employed to transfer heat between exhaust oxygen depleted air and the fresh air, where the heat generation during adsorption raises the exhaust air temperature. As result, the fresh air needs less preheating (i.e. 1020ºF) for the CAR process, compared to 1650ºF for the OITM process. On the other hand, the desorption process is associated with heat absorption, where heat has to be transferred to the bed during oxygen release. This is done by direct injection of natural gas into bed to combust with oxygen and by pre-stored heat in the sorbent bed during adsorption. The CAR process is so designed that it also stores the energy in the bed during the heat release in the adsorption cycle and releases this heat during the stripping cycle, by means of installation of inert layers on both ends of sorbent bed for heat storage. In this so called “auto-thermal recovery”, more heat is stored in bed, less heat is carried out by oxygen-depleted air, and less heat is required from fuel combustion to strip oxygen.

High oxygen concentration can be obtained by steam striping providing that the steam is condensed out downstream, and only limited oxygen concentration, about 30-40%, can be achieved if the flue gas is applied as a sweep gas. As reported in the BOC study [5] with steam as sweeping gas, the air to steam molar ratio is about 2.66 with O_2 recovery over 90%, which leads to an oxygen concentration in the sweeping gas of about 33-36%v before steam condensing.

Simulated integration of the CAR process to O_2-PC has been reported by BOC to determine the technical and economical feasibility [5]. The air-fired reference plant is an existing ultra-supercritical lignite-fired 865 MWe Lippendorf power plant near Leipzig, Germany. A simulation of an oxyfuel power plant with cryogenic air separation was employed for comparison. The same plant was then applied for integration study with CAR process, where low-pressure steam extracted from steam turbine was applied as a sweep gas. Table 7 summarizes the key results.

Table 7 - Comparison for CAR with Cryogenic ASU

Case	Air	ASU	CAR
Net Power, MWe	865	687	726
Net efficiency, %(LHV)	42.6	33.3	34.0
Efficiency drop, % point	-	9.3	8.6

Comparing the CAR to the cryogenic process, the efficiency drop reduced from 9.3 to 8.6% points, and the net power increased from 687 to 726 MWe. As reported, it is clear from the results of study that the steam consumption is a critical variable for this option. The steam extraction required for oxygen stripping is about 200 kg/s in comparison with the main steam flow as 692 kg/s.

The reason recycle flue gas was not used as sweep gas was that the flue gas has to be cleaned up to avoid any contaminates to the sorbent bed, and the effects of contaminates have not yet been studied in detail. In general, to be economic, recycle gas clean up should be avoided because the clean gas will be sent back to boiler, where it mixes with combustion gases to become dirty gas again. Optimally, the gas clean up process should be applied to the flue gas exiting from system and flowing to the CO_2 plant. Moreover, this stream has much less flow to be treated than does the recycled flue gas.

Since the CAR process is based on swings in the partial pressure of oxygen, it will be affected by pressure, and LMPD (logarithmic mean pressure difference). Increased adsorption pressure will enhance oxygen adsorption from air, and a low desorption pressure will favor adsorbed oxygen release. However too high a pressure will require more power for air compression. If steam is used as a sweep gas, heat carried by the sweep gas can be recovered through feedwater heating during steam cooling and condensing. If flue gas is used as a sweep gas, hot sweep gas can be directly fed to boiler.

In order to explore the advantage of integration with CAR process for O_2-fired combustion, an estimation has been made for the CAR process operated with recycled flue gas sweep gas. The difference in principle between steam and flue gas sweep gases is their pressure ratios, where steam extracted at 1.6 bar could continuously expand to a pressure as low as 0.038 bar to generate more power with a pressure ratio of 1.6/0.038, while the flue gas has to be compressed from about 1.0 bar to 1.6 bar to consume power with pressure ratio as 1.6/1.0. Thus, the substitution of flue gas by extracted steam reduces power because the low pressure ratio of the flue gas is replaced by the high pressure ratio of steam for the same amount of gas volume flow. The reduced power can be calculated by difference between power from steam expansion and power for recycle gas compression, without including the changes in auxiliary power and low grade heat integration. The net result is shown by Table 8.

Table 8 – Comparison for CAR Using Different Sweep Gases

Oxidant	Air	O2		
Air Separation Method	-	Cryo	CAR	CAR
Sweep Gas	-	-	steam	flue gas
Net Power, MWe	865	687	726	767
Net Efficiency, % (LHV)	42.6	33.3	34.0	35.9
Efficiency Drop, % point	-	9.3	8.6	6.7

As can be seen in Table 8, the system efficiency increases about 1.9% points when steam is replaced by recycled flue gas as the sweep gas. As compared with the cryogenic ASU, the CAR process with gas recycle sweep gas has an increased system efficiency of 2.6% in points, which is close to 3.2% points achieved by the OITM. It is obvious from the standpoint of system efficiency that the future of the CAR process is to use recycle flue gas as a sweep gas.

Note that in the CAR process, natural gas is fired to provide a portion of the heat for stripping out the adsorbed oxygen from sorbent bed, which is similar to the gas absorption-regeneration cycle. Because of auto-thermal recovery process, the ratio of energy input from the natural gas to coal is only about 3.2%, which is much less than the heat requirement by the OITM process.

4.6 Comparisons

CO_2 cannot be captured and sequestrated without incurring an energy penalty because of the potential energy stored in the pressurized liquid CO_2. A minimum of 40 kWh/klb_{CO2} additional auxiliary power is required for CO_2 compression. The difference between technologies lies in the difference in power requirements of the different CO_2 or O_2 separation techniques, and the process gain when advanced power generation is integrated, such as power generation from OITM through hot gas expansion.

Table 9 shows that compared to CAR, the OITM results in higher system efficiency and significantly more in power because of the process gain of the hot compressed air expanding through the gas turbine (gains are in reference to the cryogenic ASU O_2-PC). When the OITM technology is integrated with O_2-fired combustion, a conventional Rankine cycle power plant is upgraded to a combined cycle power plant. This improvement makes the OITM technology attractive for economic CO_2 removal.

Table 9 - Gains from OITM and CAR Compared to Cryogenic ASU

	Efficiency gain (% points)	Power gain (MW)
OITM	3.2	117
CAR/steam sweep gas	0.7	20
CAR/flue gas sweep gas	2.6	50

The OITM faces more technical challenges than does the CAR process because it operates under higher pressure and temperature. In addition to the OITM development itself, the integration and design of the boiler air heater presents a challenge (this will be explored in Task 3). An alternative to the furnace air heater is to provide the air heating by a natural gas duct burner, although the high heat required (the ratio of heat absorbed by OITM to heat input by the coal is 22%) may preclude duct firing. For the CAR process, the heat input from natural gas is only 3.2% to the total energy input.

As listed in literature, the CO_2 can be removal by means of:

Post-combustion capture - Amine process or other
Pre-combustion capture - IGCC
Oxygen-combustion – cryogenic ASU, OITM, CAR

Table 10 compares the efficiency drop and specific power requirement of the various CO_2 removal technologies. Similar to the O_2-PC designs employing cryogenic ASU and OITM, the O_2-PC CAR power plant is designed to include liquid CO_2 pumping, wet-end heat recovery, increased flame temperature, and hot gas recycle.

Table 10 – Comparison of CO2 Removal Technologies

Boiler Type		PC				IGCC
Removal Technique		post comb.	O2 fired			pre comb.
			cryo. ASU	OITM	CAR	
Efficiency drop	% ponts	11.6	6.5	3.3	4.7	6.1
CO2 removal penalty	kWh/klbCO2	188	99	42	74	98

From Table 10, it is clear that the O_2-fired PC integrated with advanced oxygen separation technology has significant advantages over both the post combustion and pre-combustion CO_2 techniques since the separation of oxygen from air is a physical process and involves less energy than the chemical separation of CO_2 from flue gas or syngas.

5.0 Conclusion

A conceptual design of a CO_2 sequestration-ready oxygen-based 460 MWe supercritical PC boiler plant was developed with integration of advanced oxygen separation techniques, such as OITM and CAR. The optimized OITM O_2-fired design case has a CO_2 removal specific power penalty of 42 kWh/klbCO_2 and a system efficiency of 36.1% compared to the air-fired system efficiency of 39.5%. Considering that CO_2 compression itself consumes 40 kWh/klbCO_2, the OITM integration into the O_2-PC is a breakthrough in CO_2 removal. The CAR process efficiency loss and specific power penalty lies approximately midway between the cryogenic ASU and OITM.

The O_2-fired PC CO_2 removal penalty with integration of OITM is nearly a quarter of that from post combustion CO_2 removal technologies, and only a half of IGCC. OITM faces significant challenges with respect to the manufacture and stability of membranes, and scale up and design of large plants.

This study will continue with the following subsequent tasks:

Task 3: Furnace and HRA Design and Analysis
Task 4: Cost Estimate

6.0 References

1. White, J.S., Buchanan, T.L., Schoff, R.L. (Parsons); Stiegel, G. (DOE); Holt, N.A., Booras G. (EPRI); Wolk, R. (WITS), "Evaluation of Innovative Fossil Fuel Power Plants with CO_2 Removal", EPRI Report No. 1000316, December 2000.

2. Wilkinson, M., Boden, John (BP), Panesar, R. (Mitsui Babcock), Allam, R. (Air Products), "CO_2 Capture Via Oxygen Firing: Optimisation of a Retrofit Design Concept for a Refinery Power Station Boiler", First National Conference on Carbon Sequestration, May 15-17, 2001, Washington DC.

3. Seltzer, Andrew and Fan, Zhen, Conceptual Design of Oxygen-Based PC Boiler, DOE # DE-FC26-03NT41736, September 2005

4. Philip A. Armstrong, E.P. Foster, Dennis A Horazak, Harry T Morehead, VanEric E. Stein, "Ceramic and Coal: ITM Oxygen for IGCC", The 22nd Pittsburgh Coal Conference, Sept.11-15, 2005, Pittsburgh, USA

5. Divyanshu Acharya, Krish R. Krishnamurthy, Michael Leison, Scott MacAdam, Vijay K. Sethi, Marie Anheden, Kristin Jordal, Jinying Yan, "Development of a High Temperature Oxygen Generation Process and Its Application to Oxycombustion Power Plants with Carbon Dioxide Capture", ibid

6. "Greenhouse Gas Emissions Control By Oxygen Firing In Circulating Fluidized Bed Boilers: Phase 1 – A Preliminary Systems Evaluation", Alstom Power Inc. Power Plant Laboratories, Prepared for NETL, PPL Report No. PPL-03-CT-09, May 15, 2003.

7. Philip A. Armstrong, "Method for Predicting Performance of an Ion Transport Membrane Unit-Operation", Air Products and Chemicals, Inc.

7.0 Bibliography

N/A

8.0 List of Acronyms and Abbreviations

ASU	Air separation unit
CAR	Ceramic auto-thermal recovery
CDT	Compressor discharge temperature
CFD	Computational fluid dynamics
FW	Foster Wheeler
GT	Gas turbine
FD	Forced draft
FGD	Flue gas de-sulfurization reactor
HHV	Higher heating value
HIPPS	High performance power system
HRA	Heat recovery area
HRSG	(Gas turbine exhaust) heat recovery area
ID	Induced draft
IGCC	Integrated gasification combined cycle
LHV	Lower heating value
LMPD	Log mean pressure difference
LMTD	Log mean temperature difference
LP	Low pressure
NOx	Nitrogen oxides
OITM	Oxygen ion transport membrane
PC	Pulverized coal
PFBC	Pressurized fluidized bed combustion
PSA	Pressure swing absorption
RH	Reheater
SCR	Selective catalytic reactor
SH	Superheater
ST	Steam Turbine
SOx	Sulfur oxides
T	Temperature
TEG	Triethyleneglycol
TET	Turbine exit temperature
UBC	Unburned carbon loss